Hydrodynamics study of Conventional (cylindrical) fluidized bed at ambient Temperature

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In partial fulfillment of the requirements

of the degree of

Bachelor of Technology (Chemical Engineering)

By

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CERTIFICATE

This is to certify that the work in this thesis entitled "Hydrodynamics study of Conventional (cylindrical) fluidized bed at ambient Temperature " submitted by Yayati Kishore Mohanta (109CH0080) in partial fulfillment of the requirements of the prescribed curriculum for Bachelor of Technology in Chemical Engineering Session 2009-2013 in the department of Chemical Engineering, National Institute of Technology, Rourkela is an authentic work carried out by him under my supervision and guidance. To the best of my knowledge the matter embodied in the thesis is his bona fide work.

Supervisor

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ABSTRACT

Now a days , the fluidized bed has got an vital application in the industial world . It is very much necessary to have the idea about the particle shape, type of distributor , minimum fluidization velocity, voidage , sphericity of particle etc in order to achieve the proper functioning of fluidized beds. In this present work, hydrodynamic of fluidized bed with regular particles was studied. The hydrodynamic behavior like pressure drop, minimum fluidization velocity accounting with different parameters were determined and they compared with the theoretical value.

The present work is been focussed on understanding the hydrodynamic behaviour in a gas-solid two phase fluidized bed. Experiments were performed in a fluidizing cylinder of height 1mtr with internal diameter 50mm . Trial experiments were performed with 2.18 mm glass beads to check proper functioning of the fluidized bed system. Hematite powder of size 10 micrometer , sand particle ranging from 0.2 mm to 0.73 mm and dolomite particle from 0.03mm to 0.53mm are used as the solid phase. The fluidization operation had been carried out with air as gas phase and the hematite powder, sand and dolomite in the fluidizing cylinder . Superficial velocity of gas has been varied in the range of 0 to 0.8 m/s . The static bed heights of the solid phase in the fluidized bed used are taken from 2cm to 8.5 cm . Each of these aspects needs to be properly studied so that we can clearly understand the nature of fluidized bed. The first of this steps is commonly the rate-controlling one, and it depends on the hydrodynamic behavior of the bed. So it's very important to know the bed hydrodynamics, and the bed expansion, pressure drop, the phase hold-ups and minimum fluidization velocity. The behavior of bed pressure drop with superficial velocity were noted and the extensive graph had been analysed. Various parameter like particle size , bed height were taken into consideration for studying the proper hydrodynamics . The values were compared with theoretical values and the error was also found. The result have been found pretty acceptable .

Keywords : Fluidization, Distributors ,Hydrodynamics, Pressure drop, Minimum fluidization velocity, Regular particles

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CONTENTS

NOMENCLATURE

H–Fluidized Bed Height, cm

H_1-Minimum height of fluid in U-tube manometer, cm

H_2-Maximum height of fluid in U-tube manometer , cm

ΔH-Manometer fluid height difference, cm

H_s- Static Bed Height, cm

ε_g- Gas holdup

ε_s- Solid Holdup

$\varepsilon, \boldsymbol{\epsilon}_{mf}$- Porosity

ΔP -Bed Pressure Drop, kPa

β- Bed Expansion Ratio

ρ- Density, Kg/m^3

μ- Viscosity of air , Kg/m-s

V- Superficial Gas velocity, m/s

U_{mf}- Minimum fluidization velocity, m/s

D_p- Diameter of particle, mm

Re-Reynold's Number, unitless

Ar-Archmedis's number , unitless

Φ, Φ_s-sphericity of the particles

ρ_s-Density of the fluidizing particles(solids)

ρ_g-Density of the fluidizing medium(air)

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CHAPTER-1

INTRODUCTION

Most of the gas-solid fluidization behavior studies that have been reported are for straight cylindrical or columnar fluidized beds. Fluidized beds have found wide applicability in many industrial processes such as, waste water treatment (Shi et al.,1984) immobilized biofilm reactions , incineration of waste materials, coating of nuclear fuel particle, crystallization, coal gasification, roasting sulfide ores (Peng and Fan,1997) and food processing (Depypere et al.,2005) etc. Cylindrical fluidized beds are very useful for fluidization of materials with a wide particle size distribution, as well as for exothermic reactions (Kim et al., 2000). They can be operated smoothly without any instability i.e. with less pressure fluctuations (Shi et al., 1984)and also for extensive particle mixing (Schaafsma et al. 2006).

1.1 Types of fluidization

Fluidization is mainly two types i.e. Particulate fluidization and aggregative fluidization. Particulate fluidization occurs when solid and fluid density difference is not much and the solids are of smaller in size. In this type fluidization the fluid velocity required to fluidize the bed is not much (example- liquid-solid systems).

Aggregative fluidization occurs when the solid and fluid density difference is more and solids are larger in size. In this type fluidization the fluid velocity required to fluidize the bed is quite high (example- gas-solid systems). In this case when fluidization occurs then bubbles form in between the solids because of the large particle size and high liquid velocity. These bubbles carry a little or no solid particles with them. When these bubbles rise from the bed then they eventually break at the surface of bed. The superficial fluid velocity in which fluid bubbles form is called the minimum bubbling velocity. Generally a bubbling fluidized bed is considered to be undesirable for industrial application. (Narayanan et.al, 2009)

1.2 Application of Fluidization

Fluidized bed reactors are used for many different purposes.These are still used to produce gasoline and other useful fuels, along with many important chemicals. Many of the industry produces polymers using FBR technology, such as rubber, polyethylene, vinyl chloride and styrene’s. Various utilities use Fluidized bed reactor for coal gasification, in nuclear power plants, and for water and waste treatment settlings.

Fluidized bed reactors allow for a cleaner, high efficient process than previous standard reactor technologies. In the basis of some advantages of the fluidized bed, fluidization techniques are widely used in industry for its different useful applications. Basically, it has been used as a mixing process in chemical industries, where enhanced reactions, combustion, and heat transfer rates are most necessary. In the mineral industry, this technique is used for separation of mineral particles having different physical properties. Fluidization has been extensively effective in quite a variety of industries from metallurgical roasting to , petroleum refinery ,coal conversion, agricultural and food processing, material processes and pharmaceutical processes . Extensive use of fluidization began in the petroleum industry with the development of fluid bed catalytic cracking. Three phase 3 fluidized bed have been applied successfully to many industrial processes such as in hydrogen oil process for hydrogenation and hydro-desulfurization of the residual oil, H-coal process for the coal liquefaction, and Fischer-Tropsch process. Some other applications are, turbulent contacting absorption for flue gas desulphurization, bio-oxidation process for waste water treatment, physical operation such as drying and other forms of mass transfer, biotechnological processes such as fermentation and aerobic waste water treatment, conversion of glucose to ethanol, pharmaceutical, methanol production and and mineral industries, oxidation of naphthalene to phathalic anhydride (catalytic), coking of petroleum residues (non-catalytic).

1.3 Importance of Gas-solid Fluidized bed in fluidization

Gas-solid fluidizations have found more industrial applications due to good solid-fluid mixing. The important advantages of the gas-solid fluidized bed are smooth, liquid-like flow of the solid particles. This permit a continuous and automatically-controlled operation with ease of handling and rapid mixing of solids which leads to a near isothermal condition throughout the bed. This results in a very simple and controlled operation with rapid heat and mass transfer rates between gas and particles, there by minimizing the overheating in case of heat sensitive products. Some of the important applications of gas-solid fluidized beds are in cement industries ,dairy and food processing and pharmaceutical industries for drying, cooling, coating and agglomeration. Formation of large scale bubbles during the fluidization reduces the heat and mass transfer rate which affect the output of the system.

Hence persistent effort have been made by the investigators to improve the quality of gas-solid fluidization by promoting bubble breakage and hindering the coalescence of bubbles which result in reduced bed expansion and fluctuation and better gas-solid mixing .This is the main disadvantage when the fluidized bed are used in chemical and petrochemical industry where it fluidization is accompanying with chemical reaction. Reducing in gas-solid mixing reduces the rate of chemical reaction.Fluidization quality is closely related to particle intrinsic properties such as particle size, particle density ,size distribution of particle and its surface characteristics. As the particle size decreases the cohesive force (i.e. Vander Wall Force) for the particle increases. So due to this effect the method of fluidization for fine particle becomes much more difficult as compared to the larger size particle. It was also pointed out by Geldartin his classification map, fine particle in Group C (small particle size and low particle density) fluidize poorly due to their strong inter-particle cohesive force, exhibiting the channeling effect , lifting like plug and forming ‘‘rat holes’’ when aerated. Therefore, the development of the reliable techniques to improve the fluidization of cohesive fine powders is required. Several external devices are based on using vibration and mechanical agitation have been suggested to improve the flow ability of cohesive fine particle.

Despite the use of these devices, handling of fine particle is still extremely difficult and wet processing, such as coating and the granulation of fine particle has been regarded as nearly impossible . Preconditioning methods have been developed that help to decrease cohesion, which results in non-bubbling fluid-like fluidization of fine powders. For example, xerographic toners (powders of micron sized particles), with cohesion can be reduced by surface coating, have been shown to transit from the solid-like regime to uniform non-bubbling fluidization as the gas flow is increased

. It has been found that the fluidization behavior of Geldart group A particles is strongly dependent on the type of fluidizing gas. In the work of Geldart and Abrahamsen (1978), it has been demonstrated that the bubbling and bed expansion behaviours of group Aparticles of alumina and glass beads greatly differ with the type and pressure of the fluidizing gas. It has been observed that a group A powder may behave more "A-like" when using different types of fluidizing gas. According to Piepers et al. (1984), physically adsorbed gases may enhance the cohesion of the particles, resulting in an increased elastic modulus of the fluidized bed and a growth in the bed expansion had been counter reacted.

1.4 ADVANTAGES OF FLUIDIZATION

The main advantage of fluidization are that the solid is vigorously agitated by when the fluid passing through the bed, and the mixing of the solid certifies that there are practically no temperature gradients in the bed even with quite exothermic or endothermic reactions. The smooth, liquid-like flow of particles allows continuous automatically controlled operations with ease of handling. The quick mixing of solids leads to nearly isothermal conditions throughout the reactor; hence the operation can be controlled simply and reliably. The circulation of solids between two fluidized beds makes it possible to transport the vast quantities of heat produced or needed in large reactors. Heat and the mass transfer rates between gas and particles are high when compared with other modes of contacting . The rate of heat transfers between a fluidized bed and an immersed object is high; hence heat exchangers require relatively small surface areas within fluidized bed reactor.

Fluidized bed has proved to be better than conventional reactors and has got many advantages like.

- It can maintain uniform temperature.
- Bed plugging and channeling can be reduced due to the movement of solids.
- Pressure drop is less due to which pumping cost can be reduced etc.
- Flexibility of better mixing.
- It can use fine catalyst particles, which minimizes the intraparticle diffusion.

However there are some disadvantages to fluidized bed like entrainment and carryover of particles, due to particle motion attrition of catalyst can occur.

1.5 DISADVANTAGES OF FLUIDIZATION

However there are also some disadvantages of fluidized bed like, catalyst attrition due to particle motion, entrainment and carryover of the particles, relatively larger of reactor size compared to fixed beds due to bed expansions, catalyst-fluid contacts per unit volume is reduced due to bed expansion, low controllability over product selectivity for complex reactions and loss of driving force due to back mixing of particles in case of transfer operations. The other disadvantages to fluidized bed such as: the difficult-to-describe flow of gas, with its large deviation from plug flow and the bypassing of the solids by bubbles, represents an inefficient contacting system. The rapid mixing of solids in the bed leads to non-uniform residence times of solids in the reactor. Friable solids are pulverized and entrained by the gas. For non-catalytic operations at high temperature the agglomeration and sintering of fine particles can necessitate a lowering in temperature of operation and reducing the reaction.

1.6 Terms related to Fluidization Phenomena

1) Minimum Fluidization Velocity (U_{mf}) – The minimum superficial velocity at which the bed just gets fluidized. At this velocity weight of the bed just gets counterbalanced by the pressure of the fluid.

2) Bed Pressure Drop (ΔP) – Measures the total weight of the bed in combination with the buoyancy and phase holdups.

3) Gas Holdup (ε_g) – It measures the volume fraction of gas. It is the ratio of volume of gas to the total volume of bed or equal to porosity.

4) Solid Holdup (ε_s) – It measures the volume fraction of solid. It is the ratio of volume fraction of solid to the total volume of bed.

6) Porosity (ε) – It is the volume occupied by gas.

CHAPTER 2

LITERATURE REVIEW

Fluidization is an established Gas-solid contacting technique. The fluidized bed can be achieved by increasing the upward velocity of the fluid through a fixed bed of solid particles. Fluidized bed technique as compared to fixed bed has the unique advantage of a smooth, liquid like flow of solid particles which allows continuous and automatically-controlled operation with ease of handling and rapid mixing of solids. In spite of these advantage, the applications of gas-solid fluidized bed have been constrained due to certain inherent drawbacks like channeling, uncontrolled bed expansion and fluctuation because of formation of bubbles and their subsequent collapsing and slugging. These not only reduce the heat and the mass transfer rate there by affecting the outcome of the system, but influences the fluidization quality to a high extent. When the particle size reduces these problems predominant in the fluidization process . A brief survey on these problem had been made in the following .

2.1 Bubbling: A gas-solid fluidized bed is characterized by the presence of gas voids or bubbles causing a resistance to mass and heat transfer. Bubbles in gas fluidized bed are very important as they are responsible for most of the features that differentiate a packed bed from a fluidized one. The quality of fluidization specially in terms of expansion and fluctuation depends largely on the formation of bubbles and their growth in the direction of flow. Modification of gas flow promotes the formation of bubbles of smaller size through the system and cause particle movement which generally results reduced expansion and fluctuation, rapid and extensive particle mixing and a consequent high heat transfer co-efficient.

2.2 Slugging: The gas-solid fluidization is characterized by the formation of bubble. The size of the bubble increases and sometimes even its diameter may become equal to that of the column. When the bubble diameter approaches the column diameter, it is termed as slugging. An aggregatively fluidized bed in a column of small diameter operated at sufficiently high gas velocity will show continuous slugging flow. Slugging affects adversely the fluidization quality. Once slugging occurs, the portion of the bed above the bubble is pushed upwards, as by a piston. Particle rain down from the slug and the slug finally disintegrates. Periodically another slug forms and thus unstable oscillatory motion is repeated. Slugging increases the problem of entrainment and lowers the performance potential of the bed. Slugging is especially serious in long narrow fluidized beds.

2.3 Channeling: The quality of fluidization specially in term of mixing is greatly affected by channeling effect . As the flow rate through the bed of particle increases towards minimum fluidization of the bed materials, channeling may be occur. The non-uniformity in size of bed materials and poor mixing between the fluid and the particles in the bed may lead to channeling. At the onset of the formation of the channeling, the fluid tend to pass through the bed along such paths of the lower particle concentration. Channeling can also result from initial non-uniformity in the bed and tends to be accentuated by stickiness of the particle which prevents them from flowing into channeled region. Where , the fluid velocity is significant, the solid particles develop an upward movement, while in case of lower fluid velocity they go downwards. The local increase in velocity through the bed above minimum fluidization, causes the bed to locally expand, thereby altering the pressure drop through that portion of the bed. The change in local pressure drop through the distributor and the combined pressure drop of the bed and the distributor quantify the channeling. A channel tends became established if the local pressure drop through the bed-distributor system decreases with increased fluid velocity. Apart from these the parameter that are studied during a fluidization process are:

- ✓ Minimum Fluidization Velocity (U_{mf})
- ✓ Pressure Drop (ΔP)

2.4 Minimum Fluidization Velocity: When a fluid passes upwards through the interstices of a bed of solids without the slightest disturbance of the solid, and this bed is called a the fixed bed. With the further increase in the velocity of the fluid, the entire bed of the solids is suspended and behaves as if its weight is counterbalanced by the force of buoyancy. At this point, the bed of solids starts behaving like a fluid. This is called the onset of fluidization and the velocity of fluid at which it happens is called minimum fluidization velocity, which is one of the most important parameter for the design of fluidizers.There are several correlations on minimum fluidization velocity proposed by Leva ,Rowe and Henwood (1989) Narsimhan Wen and Yu (2011) , Richardson etc.. Out of which Kozeny-Carman's correlation being used over a wide range of Reynolds number is given below.

$$\frac{\Delta P}{H}=150\frac{(1-\epsilon)^2\mu\,\mathrm{Umf}}{\epsilon^3(\Phi \mathrm{Dp})^2}+1.75\frac{(1-\epsilon)\rho\,\mathrm{Umf}^2}{\epsilon^3\Phi \mathrm{Dp}} \quad \ldots\ldots\ldots\ldots \text{(Kozeny-carman's equation)}$$

Where :

Δp_b= Pressure drop across bed ;

H_s= Static bed height ;

ϵ = Voidage ;

μ= Viscosity of gas;

U_{mf}= Minimum fluidization velocity;

ρ= Density of gas;

Φ = Sphericity ;

2.5 Pressure Drop: The pressure drop across the bed is another important parameter which controls the channel and slug formation and thereby mixing of the bed material with the fluidizing fluid. At a low flow rates in packed bed, the pressure drop is approximately proportional to gas velocity upto the minimum fluidization conditions. With further increase in gas velocity, the packed bed suddenly unlocks (at the onset of the minimum fluidization condition),resulting the decrease in a pressure drop. With the gas velocities beyond the minimum fluidization, the bed expands and the gas bubbles are seen to rise resulting in non homogeneity in the bed. With the increase in the gas flow, the pressure drop should remain unchanged but due to bubbling and slugging there is always a fluctuation in the pressure drop and it increases slightly . Particularly for the coarse particles, the average total pressure drop across a slugging bed may continue to increase rate.

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CHAPTER -3

EXPERIMENTAL SETUP

The aim of the present work has been precised below:

- Hydrodynamic study of gas-solid fluidized bed with hematite dust , sand particles, different sizes of dolomite particles , bed pressure drop and minimum fluidization velocity.
- Examining the effect of the gas velocity on hydrodynamic properties studies.
- Plotting the graph between Pressure drop across the bed and superficial velocity with different bed height and for different particle size.
- Study the effect of particle size on minimum fluidization velocity.

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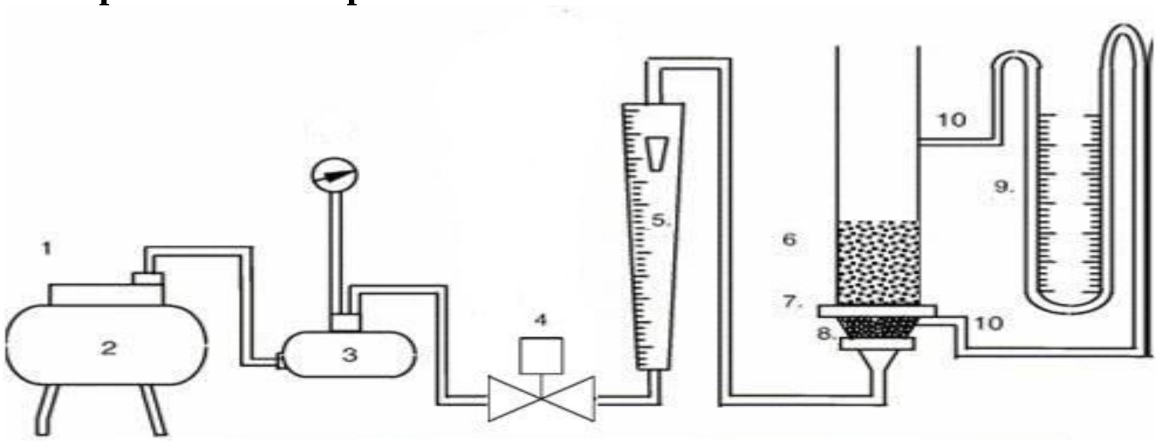
3.1 Experimental setup



<u>Experimental Setup</u>

1.Compressor	6.Fluidizer bed with bed material
2.Receiver	7.Distributor
3.Constant Pressure Tank	8.Calming section with fluidizing material
4.Valve	9.U tube Manometer
5.Manometer	10.Pressure Taping

Figure No-01 Experimental setup

3.2 Behavior of fluidized bed as superfiscial velocity increases.

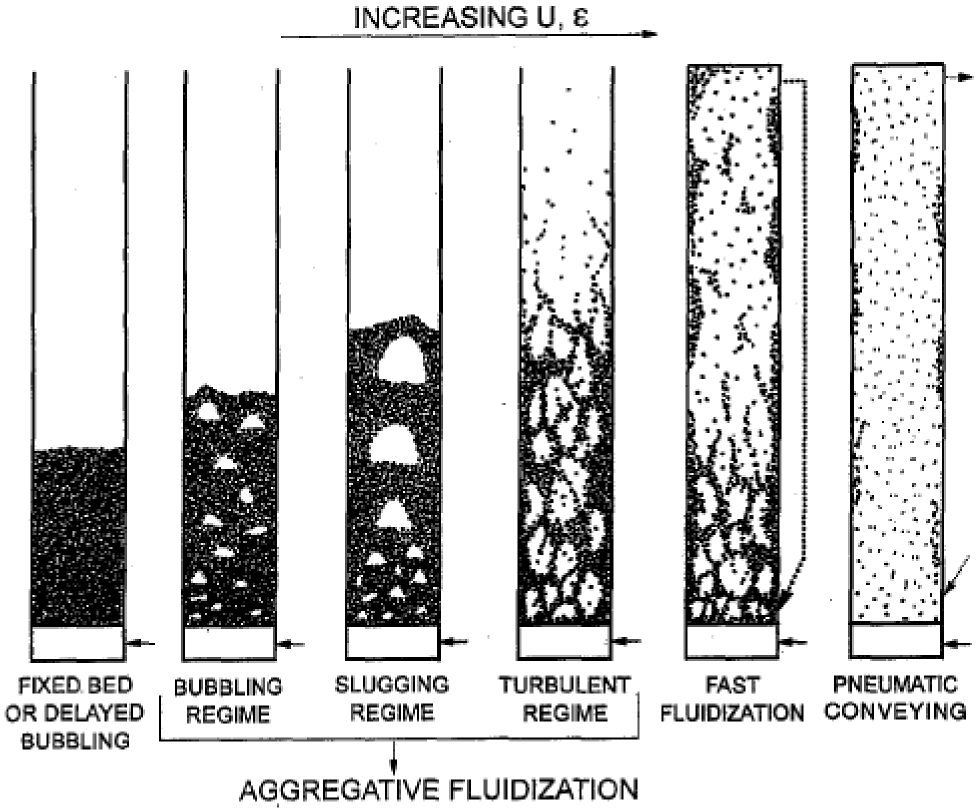


Figure No -02 Behavior of fluidized bed

3.3 Experimental Procedure

The experiment was carried put using hematite as solid particle, compressed air as gas. The gas is allowed to flow from the bottom of the fluidizing cylinder. Initially the column is filled with hematite and for different gas flow rate variation of pressure drop and bed height is taken .

Experiments were conducted at normal temperature of (30±5) C. The temperature of the air was presumed to be at normal condition ie. at 25' C. The flow rate was varied from 0 to 50 liters per minute and the readings were taken from manometer . After certain flow rate ,the bed starts to expand and the particles get fluidized. The superficial velocity was noted down. The minimum and maximum height of manometric fluid (carbon tetrachloride) were also noted down for the calculation of pressure drop across the bed.. The procedure was repeated for different gas flow rate, particles of 10 micrometer sizes and varying initial static bed heights. The same has been repeated for sand particles of size ranging from 0.20mm to 0.73 mm and dolomite particle with different bed height and different particle size ranging from 0.026 to 0.533 . The bulk density of the material of a particular size were calculated by pouring the materials into a known volume and known weight cylindrical measuring flask. The weight were noted down .The weight of the beaker was later substracted to get the bulk density.The graph between pressure drop and superficial velocity was plotted. The sphericity of the particles were calculated using Kozeny–Carman equation.Using the sphericty the minimum fluidization velocity of the particles were calculated theoretically and compared with experimental value.

3.4 OBSERVATIONS AND CALCULATIONS

3.4.1 For hematite particles(Dp=10 micro meter)

Table 01. Dp= 10 micron , Bed Height=5cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	$\Delta H = (H_2 - H_1)$ (cm)	$\Delta P = \rho g \Delta H$ (kPa)
1.25	0.01	25.2	24.0	1.2	0.187
5.00	0.04	25.3	23.9	1.4	0.218
7.50	0.06	25.5	23.6	1.9	0.296
10.0	0.08	25.6	23.5	2.1	0.327
13.75	0.12	26.0	23.2	2.8	0.436
21.25	0.18	26.0	23.3	2.7	0.420
28.75	0.24	26.0	23.3	2.7	0.420
46.25	0.29	26.0	23.3	2.8	0.420

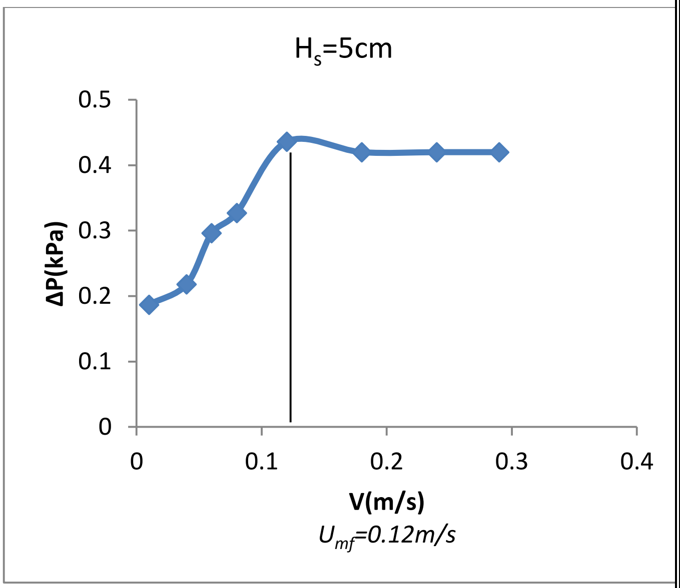


Figure 03 pressure drop vs superficial velocity

Table 02. Dp= 10 micro meter , Bed Height=6cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.5	0.01	25.1	24.0	1.1	0.172
2.5	0.02	25.8	23.2	2.6	0.407
10.0	0.08	26.0	23.0	3.0	0.467
13.25	0.11	26.3	22.8	3.4	0.529
17.50	0.15	26.0	22.6	3.4	0.529
18.75	0.16	26.0	22.6	3.4	0.529
26.50	0.22	26.0	22.6	3.4	0.529
31.25	0.26	26.1	22.6	3.5	0.545

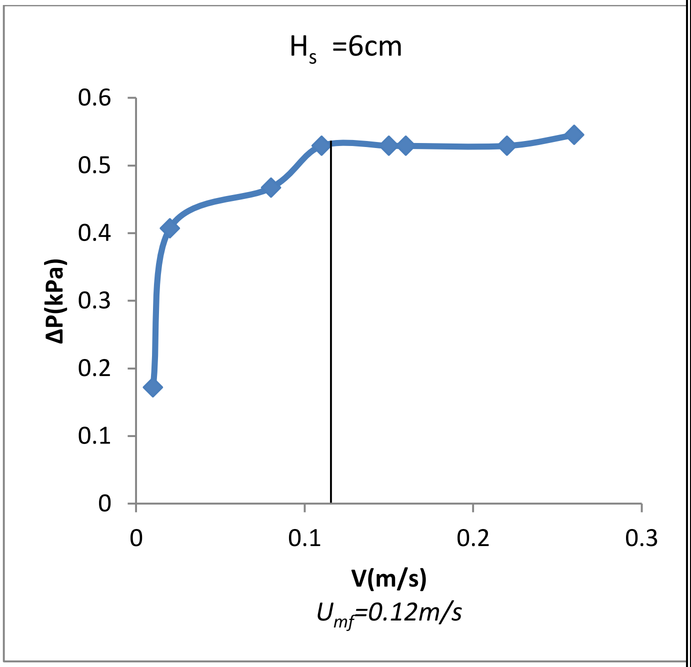


Figure 04 pressure drop vs superficial velocity

Table 03. Dp=10micron , Bed Height=7cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2 -H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	0.01	23.7	25.5	1.5	0.233
5.0	0.04	23.3	25.9	2.6	0.405
6.25	0.05	23.1	26.2	3.1	0.483
10.0	0.08	22.8	26.5	3.6	0.550
13.75	0.12	22.9	26.4	3.5	0.545
21.25	0.18	22.9	26.4	3.5	0.545
28.75	0.24	22.9	26.4	3.5	0.545
40	0.34	22.8	26.4	3.6	0.560

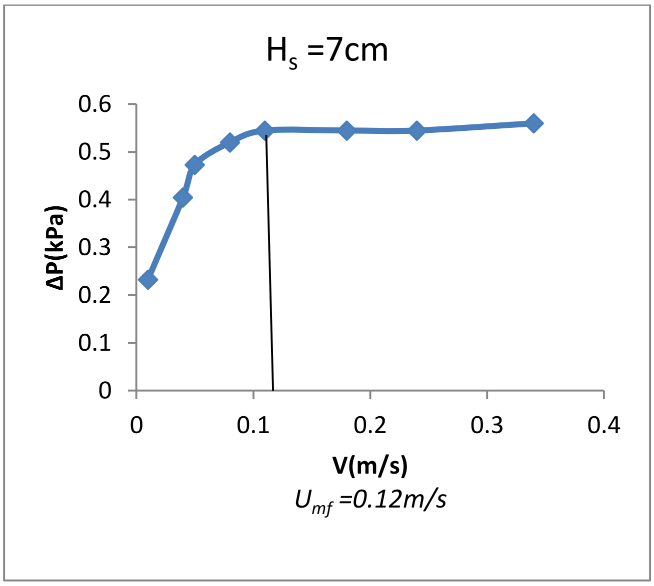


Figure 05 pressure drop vs superficial velocity

Table 04. Dp= 10micron, Bed Height=8.5cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	0.01	23.7	25.5	1.8	0.280
2.5	0.02	22.7	26.5	3.7	0.302
3.75	0.03	22.6	26.5	3.8	0.416
6.25	0.04	22.6	26.5	3.9	0.506
7.5	0.05	22.6	26.6	4.0	0.542
11.25	0.09	22.5	26.6	4.1	0.638
13.75	0.12	22.5	26.8	4.3	0.669
17.5	0.14	22.3	26.9	4.6	0.715
20.0	0.17	22.3	26.9	4.6	0.715
28.75	0.24	22.3	27.0	4.7	0.731
31.25	0.26	22.2	27.0	4.8	0.746
37.5	0.32	22.2	27.0	4.8	0.746

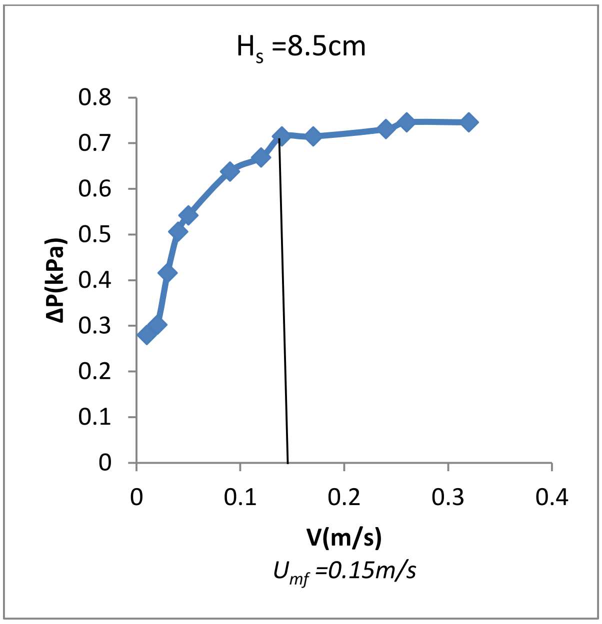


Figure 06 pressure drop vs superficial velocity

3.4.2 For Dolomite Particles

Table 05. Dp= 0.027 mm , Bed Height=6cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	0.01	26.1	23.1	3.0	0.467
2.50	0.02	26.2	23.0	3.2	0.498
3.75	0.03	26.2	22.9	3.3	0.514
5.00	0.04	26.4	22.8	3.6	0.560
6.25	0.05	26.5	22.7	3.8	0.592
10.5	0.09	26.5	22.6	3.8	0.592
11.25	0.11	26.5	22.6	3.9	0.606
21.25	0.17	26.6	22.5	4.1	0.637
40.0	0.33	26.8	22.5	4.3	0.669

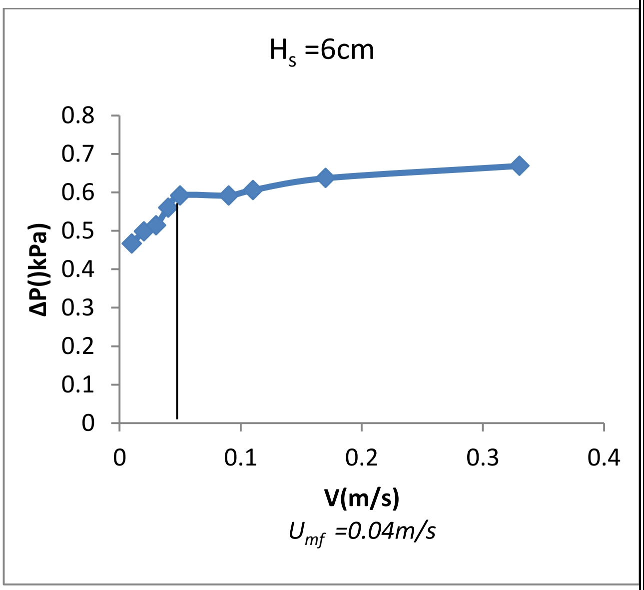


Figure 07 pressure drop vs superficial velocit

Table 06. Dp= 0.058 mm , Bed Height=2cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	**0.01**	**25.6**	**23.7**	**1.9**	**0.296**
2.50	**0.02**	**25.6**	**23.6**	**2.0**	**0.311**
3.75	**0.03**	**25.6**	**23.6**	**2.0**	**0.311**
5.0	**0.04**	**25.7**	**23.5**	**2.2**	**0.342**
6.25	**0.05**	**25.8**	**23.4**	**2.4**	**0.373**
10.25	**0.09**	**25.8**	**23.4**	**2.4**	**0.373**
11.25	**0.11**	**25.8**	**23.4**	**2.4**	**0.373**
21.25	**0.17**	**25.8**	**23.4**	**2.4**	**0.373**
40.0	**0.33**	**25.8**	**23.3**	**2.5**	**0.389**

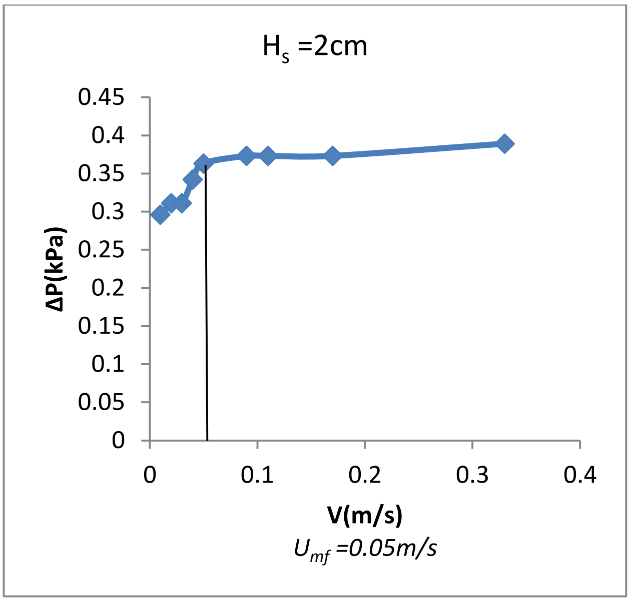


Figure 08 pressure drop vs superficial velocity

Table 07. Dp= 0.058 mm , Bed Height=3.5cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	**0.01**	**25.4**	**23.8**	**1.6**	**0.249**
2.50	**0.02**	**25.5**	**23.7**	**1.8**	**0.280**
3.75	**0.03**	**25.6**	**23.6**	**2.0**	**0.311**
5.0	**0.04**	**25.7**	**23.5**	**2.2**	**0.342**
6.25	**0.05**	**25.9**	**23.4**	**2.5**	**0.405**
10.5	**0.09**	**25.9**	**23.3**	**2.6**	**0.415**
11.25	**0.10**	**25.9**	**23.4**	**2.7**	**0.423**
21.25	**0.17**	**26.0**	**23.2**	**2.8**	**0.436**
40.0	**0.33**	**26.0**	**23.2**	**2.8**	**0.436**

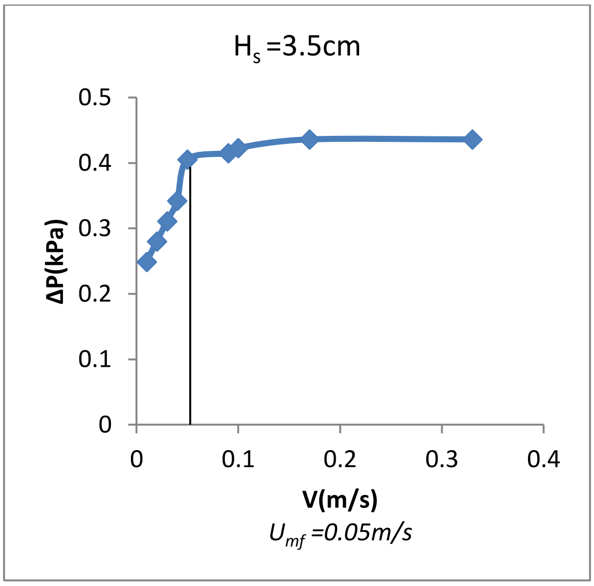


Figure 09 pressure drop vs superficial velocity

Table 08. Dp= 0.058 mm , Bed Height=4.5cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	0.01	25.5	23.7	1.8	0.280
2.50	0.02	25.6	23.6	2.0	0.311
3.75	0.03	25.7	23.5	2.2	0.342
5.0	0.04	25.9	23.3	2.6	0.405
6.25	0.05	26.0	23.3	2.7	0.420
10.5	0.09	26.0	23.2	2.8	0.436
11.25	0.10	26.0	23.2	2.8	0.436
21.25	0.17	26.1	23.1	3.0	0.480
40.0	0.33	26.2	23.0	3.2	0.501

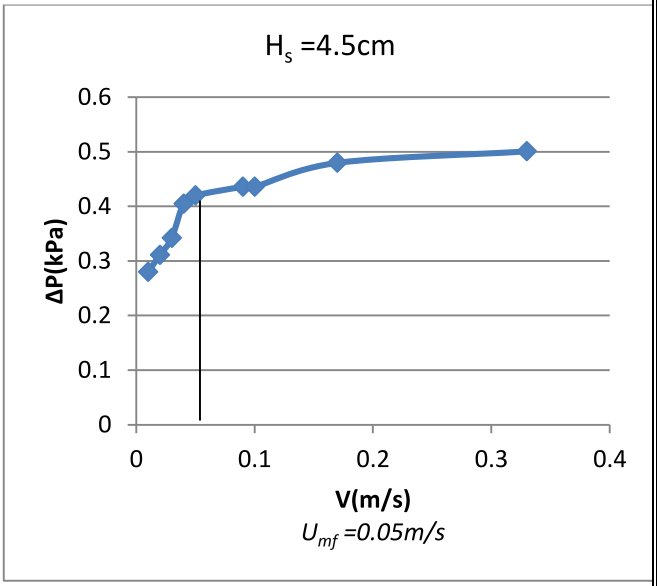


Figure 10 pressure drop vs superficial velocity

Table 09., Dp= 0.058 mm , Bed Height=6cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	0.01	25.6	23.7	1.9	0.296
2.5	0.02	25.6	23.6	2.0	0.311
3.75	0.03	25.7	23.4	2.3	0.358
5.0	0.04	25.7	23.2	2.5	0.389
6.25	0.05	26.0	23.3	2.7	0.420
10.5	0.09	26.1	23.3	2.8	0.438
11.25	0.10	26.1	23.3	2.8	0.438
21.25	0.17	26.2	23.2	3.0	0.470
40.0	0.33	26.3	23.1	3.2	0.501

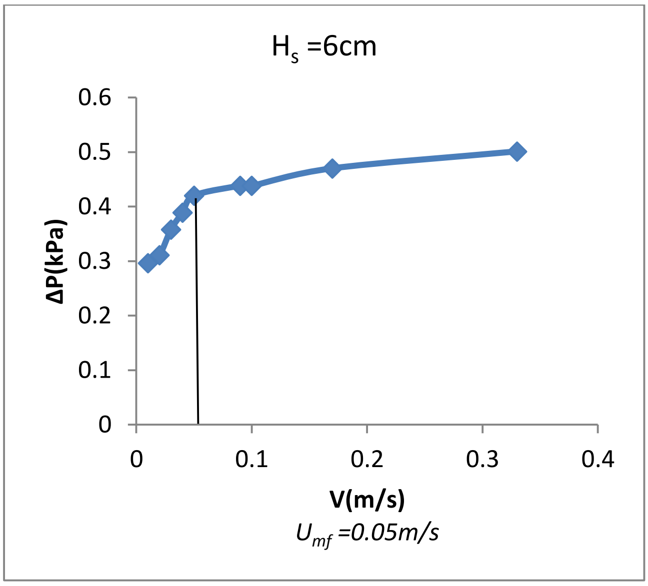


Figure 10 pressure drop vs superficial velocity

Table 10. Dp= 0.077mm , Bed Height=4.5cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	0.01	25.8	23.4	2.4	0.373
2.50	0.02	25.8	23.4	2.4	0.389
3.75	0.03	25.9	23.3	2.6	0.405
5.00	0.04	26.0	23.2	2.8	0.436
6.25	0.05	26.1	23.1	3.0	0.467
10.5	0.09	26.1	23.1	3.0	0.467
11.25	0.10	26.1	23.1	3.0	0.467
21.25	0.17	26.1	23.0	3.1	0.483
40.0	0.33	26.1	23.0	3.1	0.483

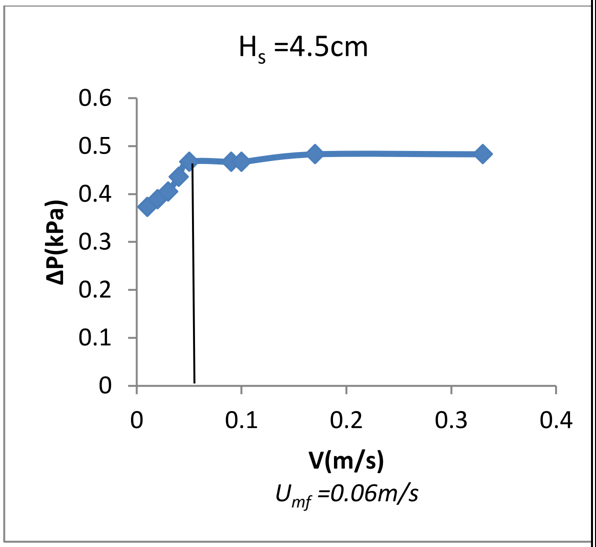


Figure 11 pressure drop vs superficial velocity

Table 11. Dp= 0.077 mm , Bed Height=3.5cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	0.01	25.7	23.5	2.2	0.342
2.50	0.02	25.8	23.5	2.3	0.349
3.75	0.03	25.9	23.4	2.4	0.373
5.00	0.04	25.9	23.3	2.6	0.405
6.25	0.05	26.0	23.2	2.7	0.421
10.5	0.09	26.0	23.2	2.8	0.436
11.25	0.10	26.0	23.1	2.9	0.452
21.25	0.17	26.0	23.2	2.8	0.436
40.0	0.33	26.0	23.1	2.9	0.452

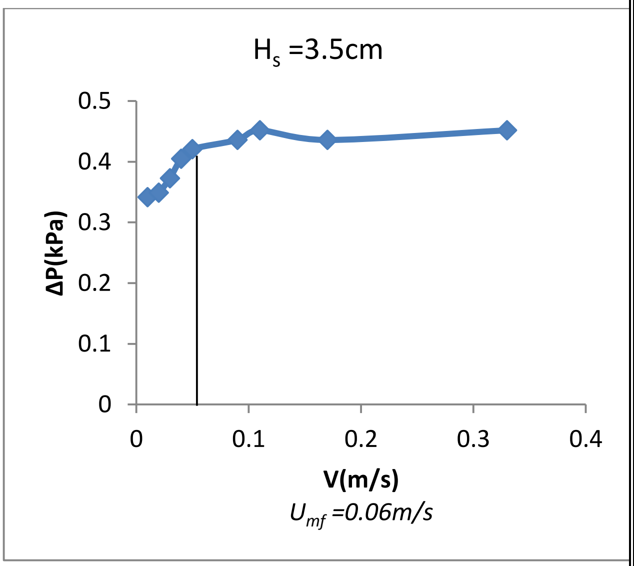


Figure 12 pressure drop vs superficial velocity

Table 12. Dp= 0.077 mm , Bed Height=2cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	0.01	25.7	23.5	2.2	0.342
2.50	0.02	25.7	23.5	2.2	0.342
3.75	0.03	25.8	23.4	2.3	0.373
5.00	0.04	25.8	23.4	2.5	0.380
6.25	0.05	25.8	23.2	2.6	0.403
10.5	0.09	25.9	23.2	2.7	0.420
11.25	0.10	26.0	23.2	2.7	0.420
21.25	0.17	26.0	23.2	2.8	0.436
40.0	0.33	26.0	23.1	2.8	0436

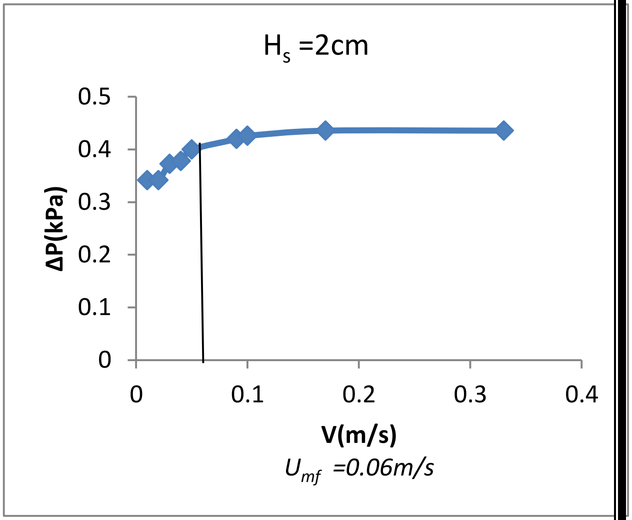


Figure 13 pressure drop vs superficial velocity

Table 13 Dp= 0.135 mm , Bed Height=6cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	0.01	25.6	23.6	2.0	0.311
2.50	0.02	26.0	23.2	2.8	0.435
3.75	0.03	26.1	23.1	3.0	0.467
5.00	0.04	26.2	23.0	3.2	0.498
10.5	0.09	26.3	22.9	3.4	0.523
11.25	0.10	26.3	23.0	3.3	0.523
21.25	0.17	26.4	23.0	3.4	0.498
40.0	0.33	26.5	23.0	3.5	0.512

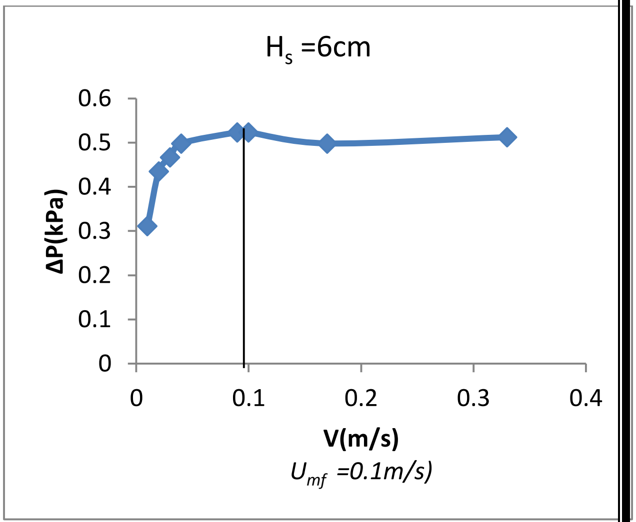


Figure 14 pressure drop vs superficial velocity

Table 14. Dp= 0.53 mm , Bed Height=6cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	0.01	26.1	23.1	3.0	0.463
2.50	0.02	26.2	23.0	3.2	0.498
3.75	0.03	26.4	22.8	3.6	0.560
5.00	0.04	26.4	22.8	3.7	0.571
6.25	0.05	26.5	22.7	3.8	0.592
10.5	0.09	26.8	22.4	4.2	0.654
11.25	0.10	26.9	22.4	4.5	0.701
21.25	0.17	26.9	22.4	4.5	0.701
40.0	0.33	26.9	22.4	4.5	0.706

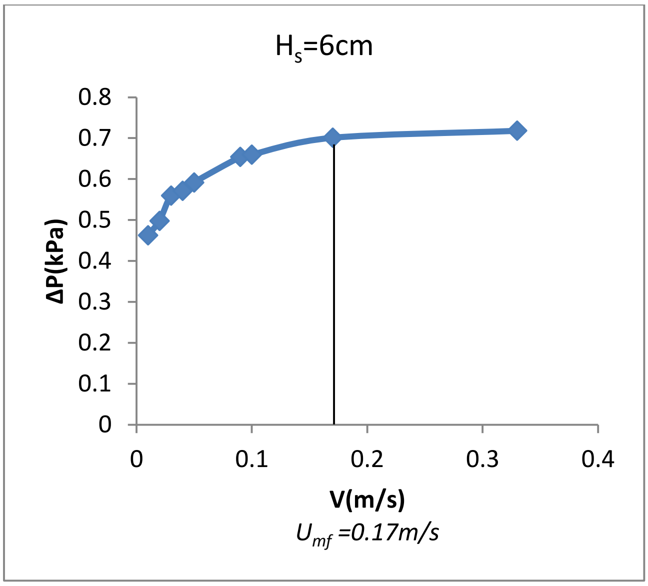


Figure 15 Pressure drop vs superficial velocity

3.4.3 For sand particles

Table 15. Dp= 0.726mm , Bed Height=6cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
2.50	0.021	22.5	21.4	1.1	0.171
11.25	0.095	22.9	21.0	1.9	0.296
13.75	0.116	23.3	20.8	2.5	0.398
18.75	0.159	23.7	20.5	3.2	0.498
23.75	0.201	24.5	19.6	3.9	0.607
28.75	0.244	24.8	19.3	5.5	0.857
36.25	0.308	25.3	18.7	6.6	1.028
45.0	0.382	25.5	18.5	7.0	1.090
52.5	0.445	25.6	18.4	7.2	1.121
55.0	0.467	25.6	18.4	7.2	1.121

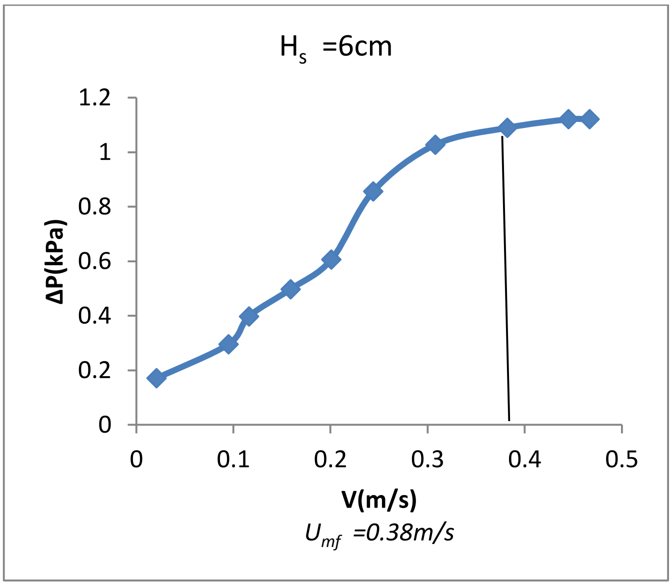


Figure 16 Pressure drop vs superficial velocity

Table 16. Dp= 0.55 mm , Bed Height=6cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
2.5	0.02	22.9	21.3	1.6	0.248
7.5	0.06	23.4	20.7	2.7	0.420
11.25	0.09	24.1	20.1	4.0	0.623
13.75	0.12	24.2	19.9	4.3	0.670
18.76	0.16	24.6	19.5	5.1	0.794
21.25	0.18	25.0	19.2	5.8	0.903
28.75	0.24	25.3	18.8	6.5	1.090
32.5	0.28	25.3	18.8	6.5	1.090
36.25	0.31	25.6	18.5	7.1	1.106
41.25	0.35	25.7	18.5	7.2	1.115

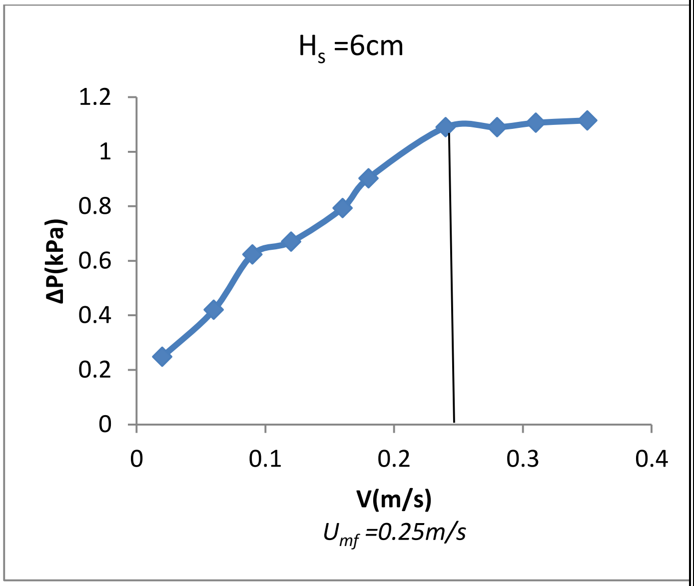


Figure 17 Pressure drop vs superficial velocity

Table 17. Dp= 0.427 mm , Bed Height=6cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
2.5	0.020	24.5	20.2	4.5	0.638
11.25	0.095	24.9	19.7	5.2	0.810
13.75	0.116	25.0	19.2	5.8	0.903
16.25	0.139	25.0	19.1	5.9	0.919
20.0	0.172	25.1	18.9	6.2	0.950
28.13	0.224	25.2	18.7	6.5	1.090
36.75	0.318	25.4	18.5	6.9	1.105
41.25	0.350	25.4	18.5	6.9	1.105

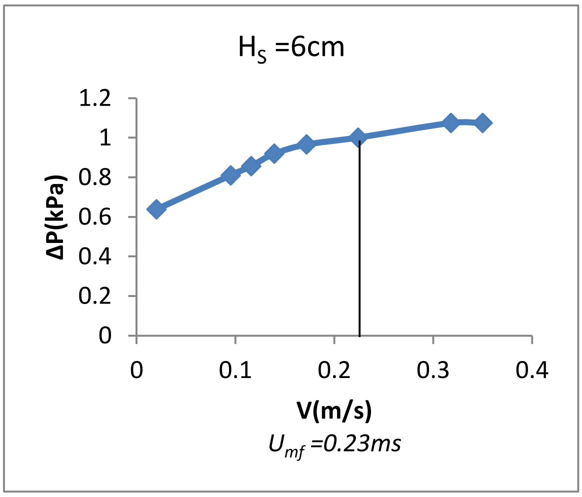


Figure 18 Pressure drop vs superficial velocity

Table 18. Dp= 0.30 mm , Bed Height=6cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
5.0	0.042	24.0	20.0	4.0	0.623
7.5	0.063	24.4	19.5	4.9	0.763
11.25	0.095	24.7	19.3	5.4	0.841
16.25	0.139	25.0	19.0	6.0	0.934
21.25	0.182	25.1	18.9	6.2	0.966
28.12	0.224	25.3	18.7	6.6	1.028
33.75	0.289	25.4	18.6	6.8	1.059

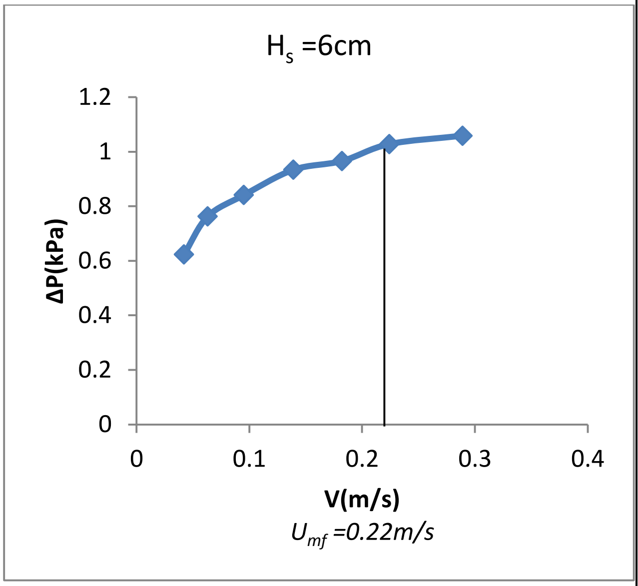


Figure19 Pressure drop vs superficial velocity

Table 19. Dp= 0.23 mm , Bed Height=6cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
2.50	0.021	24.1	20.0	4.1	0.638
7.50	0.063	24.4	19.5	4.9	0.763
16.5	0.142	24.8	19.3	5.5	0.857
21.25	0.183	25.2	19.0	5.7	0.888
16.25	0.225	25.2	18.8	6.4	0.997
36.25	0.312	25.5	18.5	7.0	1.090
40.0	0.344	25.5	18.5	7.0	1.090

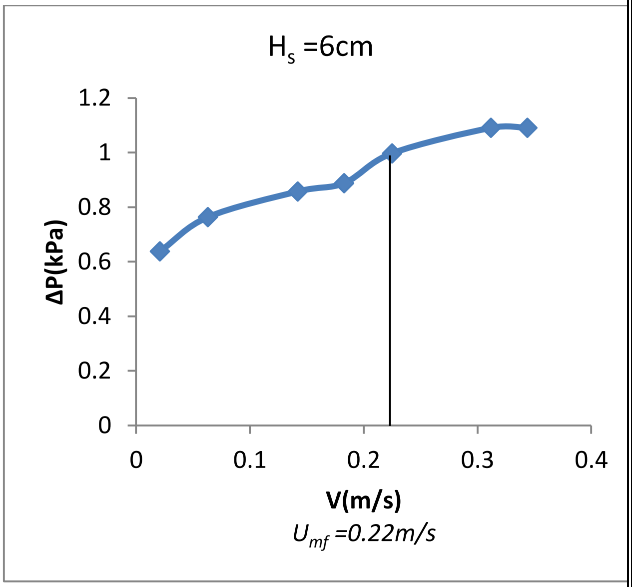


Figure 20 Pressure drop vs superficial velocity

Table 20. Dp= 0.20 mm , Bed Height=6cm

Flow rate (LPM)	Velocity (m/s)	Manometer (Minimum) H_1(cm)	Manometer (Maximum) H_2(cm)	ΔH =(H_2-H_1) (cm)	ΔP= ρgΔH (kPa)
1.25	**0.011**	**23.1**	**20.9**	**2.2**	**0.342**
2.50	**0.021**	**23.5**	**20.8**	**2.7**	**0.420**
5.00	**0.042**	**23.9**	**20.5**	**3.4**	**0.527**
11.25	**0.096**	**23.9**	**20.1**	**3.8**	**0.592**
16.25	**0.139**	**24.1**	**19.8**	**4.3**	**0.670**
18.75	**0.162**	**24.4**	**19.7**	**4.7**	**0.732**
23.75	**0.204**	**24.5**	**19.6**	**4.9**	**0.763**
26.25	**0.225**	**24.5**	**19.5**	**5.0**	**0.779**
36.25	**0.312**	**24.5**	**19.5**	**5.0**	**0.779**

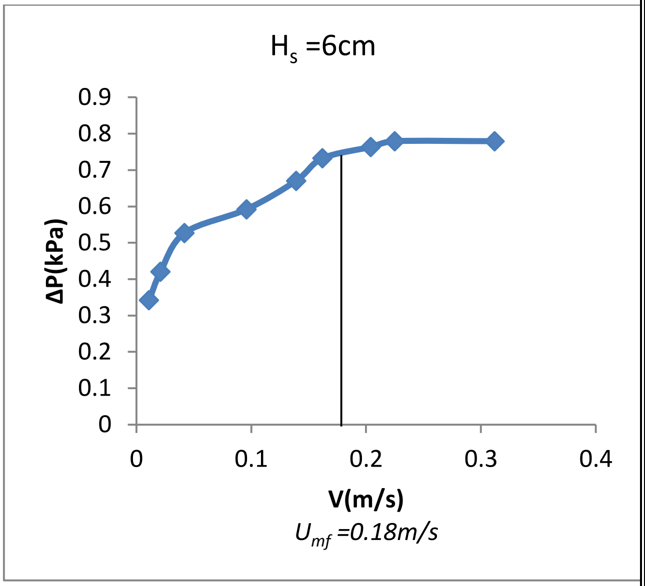


Figure 21 Pressure drop vs superficial velocity

CHAPTER-4

RESULT AND DISCUSSION

4.1 Calculation of sphericity and voidage

From Kozeny–Carman equation

$$\frac{\Delta P}{H}=150\frac{(1-\epsilon)^2\mu\ \mathrm{Umf}}{\epsilon^3(\Phi \mathrm{Dp})^2}+1.75\frac{(1-\epsilon)\rho\ \mathrm{Umf}^2}{\epsilon^3\Phi \mathrm{Dp}}$$

(The usual meaning of symbols and greek letters described in nomenclature)

The voidage is calculated by carefully pouring the particles in a 10ml measuring flask. The sphericity was then calculated by putting all the values in above formula .

The voidage of different sand particles are tabled below

For hematite bulk density =412.1 kg/m3 with a voidage of 0.92

For dolomite particle density=2740kg/m3

4.1.1 sphericity and voidage for dolomite particles

Table 21 Sphericity and voidage calculation of dolomite

Particle diameter (mm)	Sample + 10ml beaker (wt=7.295 gm)	Bulk density of sample (kg/m3)	Voidage = (1-bulk density/particle density)	Sphericity from calculation using Koeny-carman's equation
0.027	1793.8	1064.3	0.61	0.80
0.058	1767.6	1038.1	0.62	0.78
0.077	1849.4	1119.9	0.59	0.75
0.135	1898.6	1169.1	0.57	0.73
0.530	1996.0	1266.5	0.54	0.70

The sphericity and voidage calculated on the above table and it was observed that the bulk density increases as the particle diameter increases. On the other hand the sphericity decreases because, as the particle size increases the irregularity in the dimension increases.

4.1.2 Sphericity and voidage for sand particles

Table 22. Sphericity and voidage calculation of sand(ρ,particle=2600kg/m^3)

Particle diameter (mm)	Sample + 10ml beaker (wt=7.295 gm)	Bulk density of sample (kg/m^3)	Voidage =(1-bulk density/particle density)	Sphericity from calculation using Koeny-carman's quation
0.20	20.491	1319.0	0.49	0.54
0.23	20.083	1278.8	0.50	0.53
0.30	19.733	1243.8	0.52	0.50
0.43	20.522	1322.7	0.49	0.45
0.55	20.628	1333.3	0.48	0.41
0.73	20.796	1350.1	0.48	0.40

The sphericity and voidage calculated on the above table and it was observed that the bulk density increases as the particle diameter increases. On the other hand the sphericity decreases because, as the particle size increases the irregularity in the dimension increases

4.2 Calculation of theoretical U_{mf}

Following formulas used for calculating U_{mf}

Archmedis's No (Fluidization Engineering, Kunni Diazo)

$$\mathrm{Ar} = \frac{d_p^3 \rho_g (\rho_s - \rho_g) g}{\mu^2}$$

Reynold's Number $\mathrm{Re} = \frac{DV\rho}{\mu}$,

Reynolds No for fine particles By Wen And YU (1989)

$$\mathrm{Re} = [(33.7^2 + 0.0408Ar)]^{1/2} - 33.7$$

Ergun's equation for calculation of U_{mf}

$$u_{mf} = \frac{d_p^2 (\rho_s - \rho_g) g}{150\mu} \frac{\varepsilon_{mf}^3 \phi_s^2}{1 - \varepsilon_{mf}}, \qquad \mathrm{Re}_{p,mf} < 20$$

(Ergun's Equation for fine particles)

$$u_{mf}^2 = \frac{d_p (\rho_s - \rho_g) g}{1.75\rho_g} \varepsilon_{mf}^3 \phi_s, \qquad \mathrm{Re}_{p,mf} > 1000$$

(Ergun's Equation for large particles)

4.3 Theoritical U_{mf} and Experimental U_{mf} comparision for Dolomite

Table 23. Calculation of %ge error in U_{mf} of dolomite (ρ,particle=2740 kg/m^3)

Particle Diameter (mm)	U_{mf} (Theoritical)(m/s)	U_{mf} (Experimental) m/s	% error
0.027	0.06	0.04	-33.33
0.058	0.09	0.05	-44.44
0.077	0.10	0.06	-40.00
0.135	0.14	0.10	-28.57
0.533	0.27	0.17	-37.04

4.3.1 U_{mf} (theoretical , experimental) vs D_p for dolomite

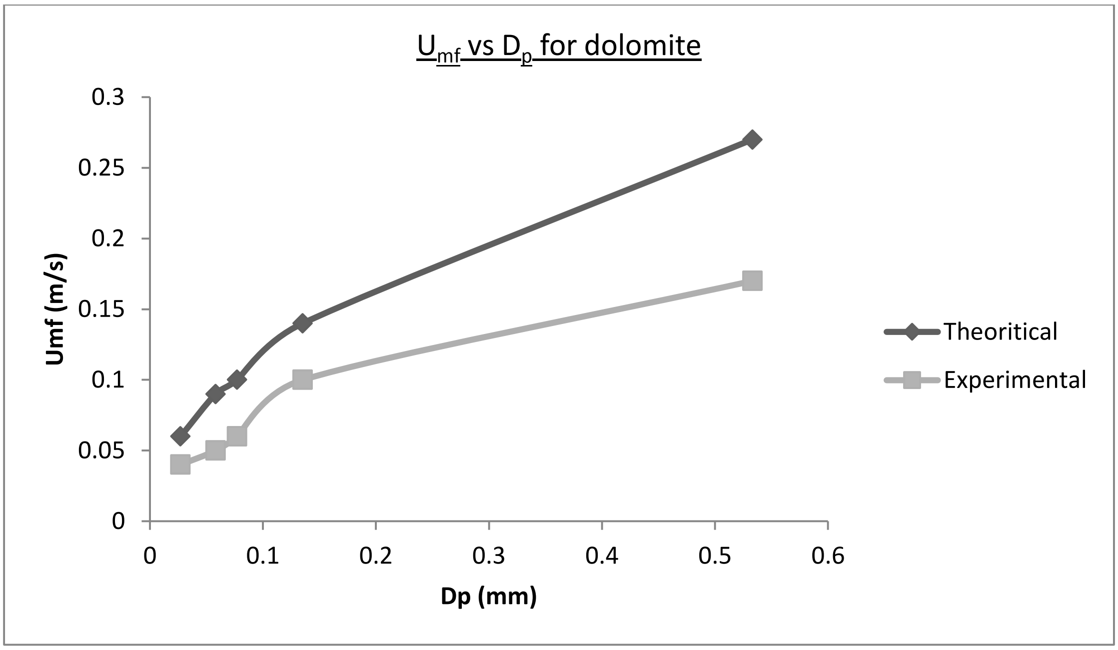


Figure 22 minimum fluidization velocity vs particle diameter for dolomite

The table 23 shows the comparision of theoretical minimum fluidization velocity with the experimental minimum fluidization velocity of the dolomite particles . In the graph (fig-22) the same is represented and it was observed that the minimum fluidization velocity increases as we increase the diameter of the particles. The experimental pattern follow the similar curve as the theoretical one . The difference in the values were calculated in terms of percentage error and all the values were got less then the theoretical value and the reason behind the same may be fault reading shown by the rota-meter , instrumental error or by leakage in the connecting air pipes , joints and valves.

4.4 Theoritical U_{mf} and Experimental U_{mf} comparision for Sand

Table 24. Calculation of %ge error in U_{mf} of sand

Particle Diameter (mm)	U_{mf} (Theoritical)(m/s)	U_{mf} (Experimental) m/s	% error
0.20	0.22	0.18	-18.18
0.23	0.24	0.22	-08.33
0.30	0.27	0.22	-18.51
0.43	0.32	0.23	-28.12
0.55	0.37	0.25	-32.43
0.73	0.42	0.38	-09.52

4.4.1 U_{mf} (Theoretical , Experimental) vs D_p for sand

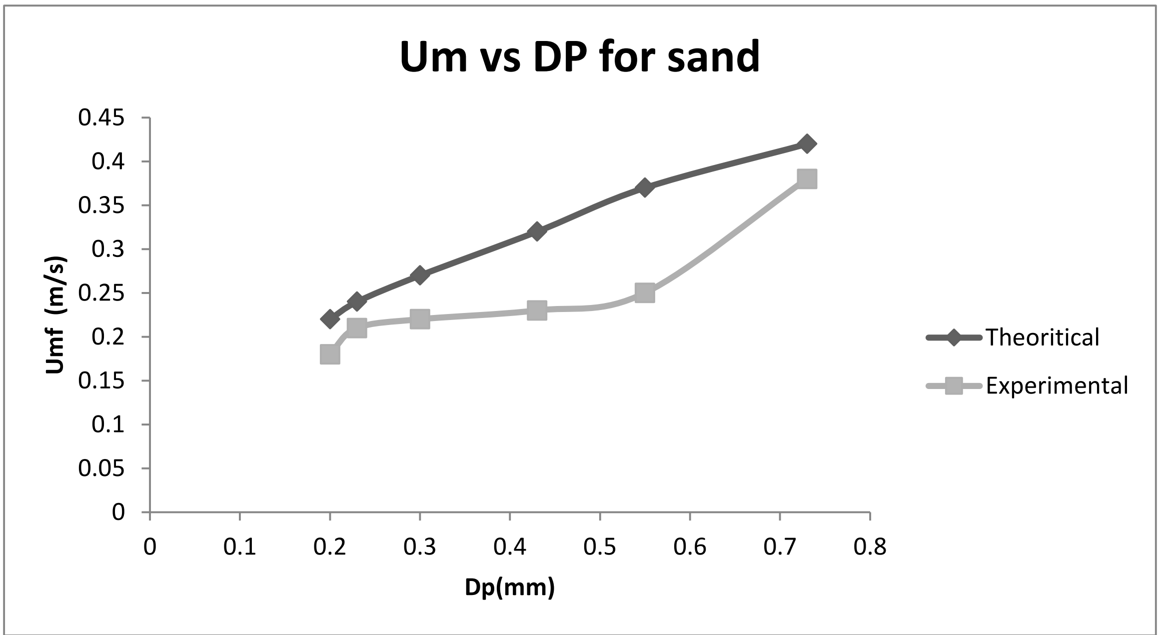


Figure 23 Minimum fluidization velocity vs particle diameter for sand particle

The table 23 shows the comparision of theoretical minimum fluidization velocity with the experimental minimum fluidization velocity of the sand particles . In the graph (fig-23) the same is represented and it was observed that the minimum fluidization velocity increases as we increase the diameter of the particles. The experimental pattern follow the similar curve as the theoretical one follows.The difference in the values were calculated in terms of percentage error and all the values were got less then the theoretical value and the reason behind the same may be fault reading showing by the rota-meter , instrumental error or by leakage in the connecting air pipes, joints and valves.

4.5 Theoritical U_{mf} and Experimental U_{mf} comparision for Hematite

Table 25. Calculation of %ge error in U_{mf} of hematite

Particle Diameter (mm)	U_{mf} (Theoritical)(m/s)	U_{mf} (Experimental) m/s	% error
0.01	0.10	0.12	20.0

4.6 Characterstics of pressure drop vs superficial velocity of different bed height for dolomite.

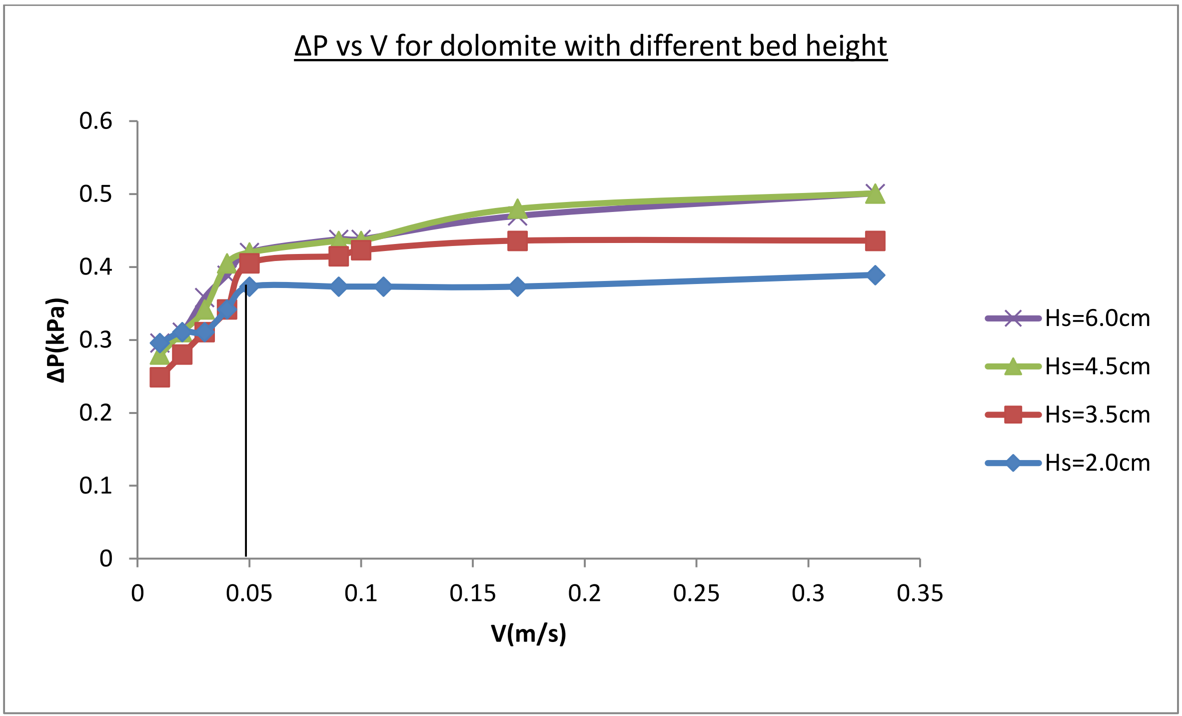


Figure 24 Pressure drop VS superficial velocity for different bed height

Fig. 24 shows the variation of pressure drop with superficial gas velocity for different bed height. The particle size for dolomite is 0.058 mm and the bed height were taken as 2cm, 3.5cm, 4.5cm and 6 cm. It was observed that the pressure drop increases with increase in static bed height. The minimum fluidization velocity was observed that, it remain unaffected with increase in the bed height.

4.7 Characterstics of ΔP vs V for same bed height , different particle size of dolomite

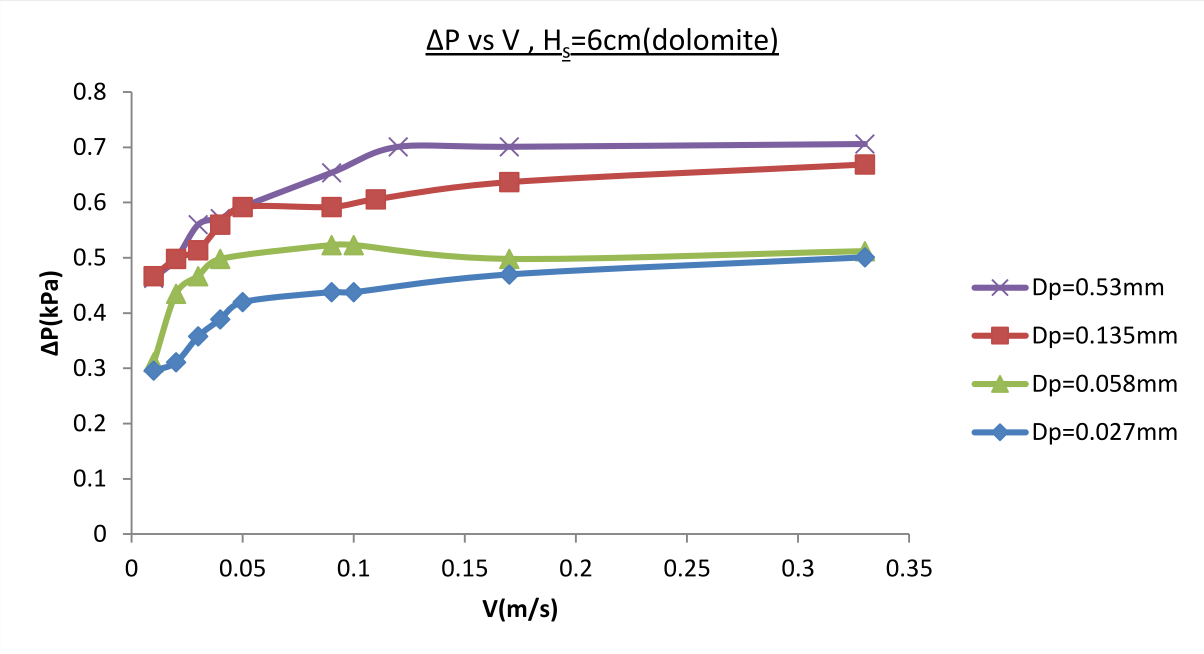


Figure 25 Pressure drop vs superficial velocity for different particle diameter , Hs=6cm

Fig. 25 shows the variation of pressure drop with gas superficial velocity for different particle size of dolomite .For particle size of 0.027 mm it attains fluidization at 0.05 m/s while the particle size of 0.058 mm it attains fluidization at 0.06 m/s .Similarly for particle diameter 0.135 the fluidization velocity achieved at 0.10m/s . For the largest particle of the experiment i.e particle size 0.53mm the fluidization velocity attained at 0.17m/s(from table 23). Thus we found that the larger size particles require much pressure force for fluidization so the pressure difference increases with increase in particle diameter. Hence minimum liquid fluidization increases with the increase in particle size.

CHAPTER-5

CONCLUSION

The graph between the superficial velocity and the pressure drop was plotted and the nature of graph was analysed .The hydrodynamic behavior of cylindrical fluidized beds is best described by the plot of pressure drop across the bed versus superficial velocity of the fluid at the entrance (bottom). Such a plot are shown in figure 3 to 21 . The graph between U_{mf} vs D_p was also plotted shown in fig no 22 , for dolomite and fig no 23, for sand. The variation of pressure drop according to different bed height was shown in fig no-24. The following conclusion were obtain

- The pressure drop increases with increase in superficial velocity upto the critical fluidization velocity and there after it relatively remain constant with a little fluctuation (fig 3 to 20)
- The pressure drop across the bed increases as we increase the bed height.(fig 25)
- The pressure drop increases with increase in particle size (fig 22 &23)
- The minimum fluidization velocity doesn’t effected by increasing bed height (fig 24).

The critical fluidization velocity and maximum pressure drop were calculated. It was also noticed that the pressure drop increases as we increase the superficial velocity. After critical velocity achieved , the pressure drop remain relatively constant after that .The minimum fluidization velocity increases as the particle size increases .The results were compared with theoretical values and error was found within a maximum of 44.44%.

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